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EP 0 432 321 B1

Description

This invention relates to an improved process for the conversion of light olefins to gasoline boiling range hydrocarbons. In particular, the invention relates to an improved technique for the recovery and separation of liquefied petroleum gas(LPG) from an olefins to gasoline conversion process effluent stream.

Conversion of olefins to gasoline and/or distillate product is disclosed in U.S. Patents 3,960,978 and 4,021,502 (Givens,Plank and Rosinski) wherein gaseous olefins in the range of ethylene to pentene, either alone or in admixture with paraffins, are converted into an olefinic gasoline blending stock by contacting the olefins with a catalyst bed made up of ZSM-5 or related zeolite. In U.S. Patents 4,150,062 and 4,227,992 Garwood et al discloses the operating conditions for the Mobil Olefin to Gasoline/Distillate (MOGD) process for selective conversion of $C_3 +$ olefins. A fluidized bed process for converting ethene-containing light olefinic streams, sometimes referred to as the Mobil Olefin to Gasoline (MOG) process is described by Avidan et al in U.S. Patent Application 006,407, filed 23 Jan 1987. The phenomena of shape-selective polymerization are discussed by Garwood in ACS Symposium Series No. 218, Intrazeolite Chemistry, "Conversion of C_2 - C_{10} to Higher Olefins over Synthetic Zeolite ZSM-5", 1983 American Chemical Society.

In the process for catalytic conversion of olefins to heavier hydrocarbons by catalytic oligomerization using an acid crystalline metallosilicate zeolite, such as ZSM-5 or related shape selective catalyst, process conditions can be varied to favor the formation of either gasoline or distillate range products. In the gasoline operating mode, or MOG reactor system, ethylene and the other lower olefins are catalytically oligomerized at elevated temperature and moderate pressure. Under these conditions ethylene conversion rate is greatly increased and lower olefin oligomerization is nearly complete to produce $C_5 +$ hydrocarbons in good yield.

The olefins contained in an FCC gas plant are an advantageous feed for MOG. U.S. Patent No. 4,090,949 discloses upgrading olefinic gasoline by conversion in the presence of carbon hydrogen-contributing fragments including olefins and a zeolite catalyst and where the contributing olefins may be obtained from a gas plant. U.S. Patent Nos. 4,471,147 and 4,504,691 disclose an MOG/D process using an olefinic feedstock derived from FCC effluent. In these two latter patents the first step involves prefractationating the olefinic feedstock to obtain a gaseous stream rich in ethylene and a liquid stream containing $C_3 +$ olefin.

US-A-4456781 disclosed a continuous catalytic process for oligomerizing lower olefins to heavier hydrocarbons. In this process a continuous liquid olefinic feedstream is diluted with a liquid alkane stream and contacted with oligomerization catalyst.

The conventional MOG process design is concerned with converting ethylene in a fuel gas stream, such as an FCC off-gas, to gasoline. In the conventional MOG design no LPG recovery facility is provided since the LPG content of the MOG reactor effluent is relatively small. However, when it is desired to convert propene and/or butene to gasoline by processing olefinic-paraffinic LPG the unreacted paraffinic LPG, unconverted olefinic LPG and LPG produced in the conversion step constitute a significant portion of the MOG reactor effluent. In this case, processing the reactor effluent in the conventional MOG design is unacceptable since a major portion of reactor effluent LPG will be lost to fuel gas. However, with an adequate recovery and separation design for the LPG content of an MOG process converting C_2 - C_4 olefins the performance of the MOG process could be improved where the process would represent a viable alternative to acid catalyzed alkylation as a route to high octane gasoline. Further, an economical recovery and separation step will open up the MOG process to utilize a wider range of available feedstock, particularly FCC light olefinic products, routinely available in the refinery setting. The provision of an improved MOG process as an alternative to the economically and environmentally beleaguered alkylation process would constitute a very noteworthy contribution to the options available to the refinery arts for the production of high octane.

The present invention provides a process wherein a fractionation step is incorporated into the recovery and separation of the effluent from the olefins to gasoline (MOG) process such that the LPG components of the effluent stream are separated and recovered as well as a stream comprising $C_5 +$ gasoline range boiling liquids. In a preferred embodiment of the present invention the effluent stream is separated in high temperature and low temperature separators and the low boiling fraction is deethanized in a conventional absorber-sponge absorber system while higher boiling component, following stripping, is passed to the depropanizing-debutanizing section of the process.

It has further been discovered that the process of the present invention can be integrated with an unsaturated gas plant debutanizer upstream of the olefins to gasoline conversion reactor. In this embodiment the feedstream to the FCC debutanizer, comprising wild gasoline and FCC wet gas is passed to the FCC debutanizer and the vapor overhead fraction therefrom is passed to the MOG reactor system. Optionally, the FCC debutanizer can be replaced with a depropanizer and a common debutanizer utilized to

separate both the MOG effluent after deethanization and depropanization and the bottoms effluent from the FCC depropanizer.

More specifically, an improved process for the conversion of lower olefinic hydrocarbon feedstock to C₅ + gasoline range hydrocarbons has been discovered comprising: contacting a hydrocarbon stream containing C₃- and/or C₄-olefinic hydrocarbons with a medium pore shape selective solid catalyst in oligomerization zone under oligomerization conditions to produce an effluent stream rich in C₅ + gasoline range hydrocarbons; separating said effluent stream to provide a C₃- hydrocarbon stream and a C₃ + hydrocarbon stream; fractionating said C₃ + hydrocarbon stream to produce a C₅ + gasoline range hydrocarbon stream, a stream rich in C₄ hydrocarbons and a stream rich in C₃ hydrocarbons.

In the drawings,

Figure 1 is a schematic flow diagram illustrating the basic process design of the instant invention.

Figure 2 is a flow diagram showing the novel MOG process integration with FCC unsaturated gas plant for the purpose of converting C₄- olefins in MOG.

Figure 3 is a process flow diagram further illustrating the novel MOG process integration with unsaturated gas plant for the purpose of converting C₃-olefins in MOG.

The present invention provides a system for upgrading light olefins such as FCC product components obtained from a FCC main column overhead product debutanizer or depropanizer, to liquid hydrocarbons. The invention utilizes a continuous process for producing fuel products by oligomerizing olefinic components to produce higher hydrocarbon products for use as fuel or the like. It provides a separation technique for use with processes for oligomerizing lower alkene-containing light gas feedstock, optionally containing ethene, propene, butenes or lower alkanes, to produce predominantly C₅ + hydrocarbons, including olefins.

The preferred feedstock contains C₂-C₄ alkenes (mono-olefin) in the range of 10 to 90 wt%. Non-deleterious components, such as methane and other paraffins and inert gases, may be present. A particularly useful feedstock is a light gas by-product of FCC gas oil cracking units containing typically 10-40 mol % C₂-C₄ olefins and 5-35 mol % H₂ with varying amounts of C₁-C₃ paraffins and inert gas, such as N₂. The process may be tolerant of a wide range of lower alkanes, from 0 to 95%. Preferred feedstocks contain more than 50 wt % C₁-C₄ lower aliphatic hydrocarbons, and contain sufficient olefins to provide total olefinic partial pressure of at least 50 kPa. Under the reaction severity conditions employed in the present invention lower alkanes, especially propane, may be partially converted to C₄ + products.

Conversion of lower olefins, especially ethene, propene and butenes, over HZSM-5 is effective at moderately elevated temperatures and pressures. The conversion products are sought as liquid fuels, especially the C₅ + hydrocarbons. Product distribution for liquid hydrocarbons can be varied by controlling process conditions, such as temperature, pressure and space velocity. Gasoline (eg, C₅-C₉) is readily formed at elevated temperature (e.g., up to 400 °C) and moderate pressure from ambient to 5500 kPa, preferably 250 to 2900 kPa. Under appropriate conditions of catalyst activity, reaction temperature and space velocity, predominantly olefinic gasoline can be produced in good yield and may be recovered as a product. Operating details for typical olefin oligomerization units are disclosed in U.S. Patents 4,456,779; 4,497,968 (Owen et al.) and 4,433, 185 (Tabak).

It has been found that C₂-C₄ rich olefinic light gas can be upgraded to liquid hydrocarbons rich in olefinic gasoline by catalytic conversion in a turbulent fluidized bed of solid acid zeolite catalyst under low severity reaction conditions in a single pass or with recycle of gaseous effluent components. This technique is particularly useful for upgrading LPG and FCC light gas, which usually contains significant amounts of ethene, propene, butenes, C₂-C₄ paraffins and hydrogen produced in cracking heavy petroleum oils or the like.

Recent developments in zeolite technology have provided a group of medium pore siliceous materials having similar pore geometry. Most prominent among these intermediate pore size zeolites is ZSM-5, which is usually synthesized with Bronsted acid active sites by incorporating a tetrahedrally coordinated metal, such as Al, Ga, or Fe, within the zeolytic framework. These medium pore zeolites are favored for acid catalysis; however, the advantages of ZSM-5 structures may be utilized by employing highly siliceous materials or crystalline metallosilicate having one or more tetrahedral species having varying degrees of acidity. ZSM-5 crystalline structure is readily recognized by its X-ray diffraction pattern, which is described in U.S. Patent No. 3,702,866 (Argauer, et al.).

The oligomerization catalyst preferred for use in olefins conversion includes the medium pore (i.e., 5-7 x 10⁻⁷ mm (Angstroms)) shape selective crystalline aluminosilicate zeolites having a silica to alumina ratio of 20:1 or greater, a constraint index of 1-12, and acid cracking activity (alpha value) of 10-200. Representative of the shape selective zeolites are ZSM-5, ZSM-11, ZSM-12, ZSM-22, ZSM-23, ZSM-35, ZSM-38, and ZSM-48. ZSM-5 is disclosed in U.S. Patent No. 3,702,886 and U.S. Patent No. Reissue 29,948. Other suitable zeolites are disclosed in U.S. Patent Nos. 3,709,979 (ZSM-11); 3,832,449 (ZSM-12); 4,076842

(ZSM-23); 4,016,245 (ZSM-35); and 4,375,573 (ZSM-48).

While suitable zeolites having a coordinated metal oxide to silica molar ratio of 20:1 to 200:1 or higher may be used, it is advantageous to employ a standard ZSM-5 having a silica alumina molar ratio of 25:1 to 70:1, suitably modified. A typical zeolite catalyst component having Bronsted acid sites may consist essentially of aluminosilicate ZSM-5 zeolite with 5 to 95 wt % silica, clay and/or alumina binder.

These siliceous zeolites may be employed in their acid forms ion exchanged or impregnated with one or more suitable metals, such as Ga, Pd, Zn, Ni, Co and/or other metals of Periodic Groups III to VIII. Ni-exchanged or impregnated catalyst is particularly useful in converting ethene under low severity conditions. The zeolite may include other components, generally one or more metals of group IV, IIB, IIIB, VA VIA or VIIIA of the Periodic Table (IUPAC). Useful hydrogenation components include the noble metals of Group VIIIA, especially platinum, but other noble metals, such as palladium, gold, silver, rhenium or rhodium, may also be used. Base metal hydrogenation components may also be used, especially nickel, cobalt, molybdenum, tungsten, copper or zinc. The catalyst materials may include two or more catalytic components, such as a metallic oligomerization component (eg, ionic Ni^{+2} , and a shape-selective medium pore acidic oligomerization catalyst, such as ZSM-5 zeolite) which components may be present in admixture or combined in a unitary bifunctional solid particle. It is possible to utilize an ethene dimerization metal or oligomerization agent to effectively convert feedstock ethene in a continuous reaction zone. Certain of the ZSM-5 type medium pore shape selective catalysts are sometimes known as pentasils. In addition to the preferred aluminosilicates, the borosilicate, ferrosilicate and "silicalite" materials may be employed.

ZSM-5 type pentasil zeolites are particularly useful in the process because of their regenerability, long life and stability under the extreme conditions of operation. Usually the zeolite crystals have a crystal size from 0.01 to over 2×10^{-3} mm or more, with 0.02-1 micron being preferred.

A further useful catalyst is a medium pore shape selective crystalline aluminosilicate zeolite as described above containing at least one Group VIII metal, for example Ni-ZSM-5. This catalyst has been shown to convert ethylene at moderate temperatures and is disclosed in commonly assigned U.S. Patent 4,717,782.

Referring now to Figure 1, the novel process of the instant invention is shown which allows the utilization of LPG streams containing propene and butene as feedstock to the MOG process in addition to fuel gas containing ethene. Feedstock is introduced to the MOG reactor by conduits 101 and/or 102. In the present embodiment the feedstock may be drawn from any refinery source. The effluent from the conversion reactor is passed 103 after cooling 104 to a high temperature separator 105 for separation of a high boiling fraction 106 containing C_5 + hydrocarbons. That fraction is passed to stripper means 107. The vapor fraction 108 from the high temperature separator is cooled 109 and passed to a low temperature separator 110 and a higher boiling component of that fraction is separated and passed 111 to stripper 107. The light fraction 112 from separator 110 comprising light hydrocarbons is passed to absorber and sponge absorber system 113 and 114 for deethanization and recovery of C_2 - off-gas 115. The overhead fraction 116 from stripper 107 is recycled to the high temperature separator and the bottom fraction 108' comprising C_4 + hydrocarbons is passed through the novel depropanizer debutanizer of the present invention 117 where bottom C_5 + MOG gasoline fraction is separated 118. Stream 119 is withdrawn from a mid-portion of fractionator 117. This more efficiently separates C_3 and C_4 components as a bottom C_4 stream and an overhead C_3 stream 121 which is recycled to a top portion of fractionator 117 for separation as an overhead stream comprising C_3 hydrocarbons 122.

An important advantage of the present invention is to be found in those embodiments wherein the downstream separation of the effluent from an MOG reactor is integrated with an existing unsaturated gas plant such as the unsaturated gas plant (USGP) commonly incorporated as part of a fluid catalytic cracking (FCC) operation. The advantages inherent in these embodiments of the present invention lie in two general directions: the ability to double up on the utilization of USGP separation towers which affords a significant economic advantage in the costs associated with separation of the MOG reactor effluent; the opportunity to down-load USGP towers by shifting deethanization, depropanization and debutanization operations in large part to the towers integrated into the design of the MOG reactor effluent separation, inherent within the present invention.

Figures 2 and 3 present process flow diagrams representative of embodiments of the present invention wherein the invention involves an integration of the MOG product separation operations with unsaturated gas plant operations in a generic way, they are illustrative of the integration of MOG product effluent separation and USGP operation. The configuration of actual integrations may vary depending upon site specific and market specific opportunities in ways which can obviously be derived from the generic embodiments presented herein by those skilled in the art.

Referring now to Figure 2, FCC wild gasoline 210 and the product outlet from the after-cooler from an FCC wet gas compressor 211 are passed to an FCC debutanizer 212 after separation of the after-cooler outlet stream into vapor and liquid components 213 and 214. A bottom stream 215 is separated from the debutanizer comprising C₅ + FCC gasoline and the overhead stream 216 comprising C₄-hydrocarbons is passed to the MOG reactor 217. Optionally, feed from other process units comprising C₂-C₄ olefins is also passed 218 to the reactor. The MOG reactor effluent 219 is cooled and separated into liquid and vapor fractions in a low temperature separator 220. The liquid portion is passed to stripper 221 and the bottom portion therefrom is passed 222 to debutanizer 223 for separation into C₄-overhead 224 and C₅ + MOG gasoline 225. Vapor from low temperature separator 220 is passed 226 to an absorber/sponge absorber system 227,228 for deethanization. A portion of stream 215 is passed 235 to absorber 227 as lean oil. The less volatile FCC gasoline stream 215 is the preferred lean oil because less volatile lean oil usage results in less gasoline carry over to the sponge absorber. The overhead from stripper 221 and the bottom fraction from absorber 227 is recycled 229 and 230 to low temperature separator 220.

Figure 3 illustrates an embodiment of the present invention integrating MOG product separation with an FCC unsaturated gas plant utilizing a common debutanizer for separation of FCC and MOG product. Referring to Figure 3, FCC wild gasoline is passed 310 to a depropanizer 311 in conjunction with the vapor and liquid fractions 312 and 313 from the FCC wet gas compressor after-cooler. C₃- overhead is passed 314 as a feed stream to the MOG reactor 315. The effluent therefrom 316 is separated via low temperature separator 317 and stripper 318 to provide a C₃ + fraction 319 and an overhead fraction 320 which is deethanized in absorber system 321 and 322. In this case the bottom stripper fraction is passed to a depropanizer 323 and a C₄ + bottom fraction 324 is separated. This fraction is passed to a debutanizer in conjunction with the bottom fraction 325 from depropanizer 311. In the common debutanizer a bottom fraction is separated 326 comprising MOG and FCC C₅ + gasoline and an overhead fraction is collected 327 comprising MOG and FCC C₄ fractions.

In the following, (Table I) a comparison is presented showing the advantages of the present invention over conventional MOG operations. Column A shows the product distribution of an unsaturated gas plant, not incorporating an MOG process unit. Column B shows the product distribution of a conventional MOG operation which uses as a feedstock treated FCC sponge absorber stream. Column C shows a product distribution from MOG and USGP integration of the present invention represented by Figure 2 process flow diagram. The results clearly show a distinctly superior yield of total gasoline product in the process of the instant invention.

TABLE I

MOG/USGP DESIGN EFFECT ON PRODUCT DISTRIBUTION			
	Column A	Column B	Column C
MOG Gasoline(BPSD*)	-	780	5212
FCC Gasoline(BPSD*)	30995	30995	30995
Total Gasoline(BPSD)	30995	31775	36207
Butene(BPSD)	4227	4163	273
i-Butane(BPSD)	1791	1840	2235
n-Butane(BPSD)	1134	1149	1150
Total Liquid C ₄ 's	7152	7152	3658
Propene(BPSD)	3775	3689	116
Propane(BPSD)	1155	1174	1191
Total Liquid C ₃ 's	4930	4863	1307
Fuel Gas(MMSCFD)	12.6	10.7	11.2

*Barrels per stream day

In Table II a comparison is presented of the equipment and energy fractionation requirements for an unsaturated gas plant alone and an integrated MOG/USGP unit. The comparison shows the advantages of MOG/USGP of the instant invention which can be operated with the same energy usage and equipment requirements as a USGP alone.

TABLE II

	Tower Diameter, m(Ft.)		Reboiler Duty (MMBTU/HR)		Condenser Duty (MMBTU/H R)	
	USGP	MOG/USGP	USGP	MOG/USGP	USGP	MOG/USGP
Sponge Absorber	1.37 (4.5)	1.22 (4.0)	0	0	0	0
Absorber/Stripper	2.59 (8.5)	1.83 (6.0)	70	27	0	0
FCC Gasoline Debutanizer	3.20 (10.5)	3.66 (12.0)	59	83	41	14
MOG Gasoline Debutanizer	0	1.83 (6.0)	0	27	0	18
Depropanizer	1.83 (6.0)	1.22 (4.0)	15	6	14	6

While the invention has been shown by describing preferred embodiments of the process, there is no intent to limit the inventive concept except as set forth in the following claims.

Claims

1. An improved process for the conversion of lower olefinic hydrocarbon feedstock to C₅ + gasoline range hydrocarbons comprising:
 - a. contacting a hydrocarbon stream containing C₃-and/or C₄- olefinic hydrocarbons with a medium pore shape selective solid catalyst in an oligomerization zone under oligomerization conditions to produce an effluent stream rich in C₅ + gasoline range hydrocarbons;
 - b. separating said effluent stream to provide a C₃-hydrocarbon stream and a C₃ + hydrocarbon stream;
 - c. fractionating said C₃ + hydrocarbon stream to produce a C₅ + gasoline range hydrocarbon stream, a stream rich in C₄ hydrocarbons and a stream rich in C₃ hydrocarbons.
2. The process of Claim 1 wherein step (b) separation of said effluent stream comprises passing said stream to high temperature and/or low temperature separators; passing the gaseous portion from said separators to absorber means whereby C₂-hydrocarbon stream is produced; and passing liquid portion from said separators to a stripping means whereby C₃ + hydrocarbon stream is produced.
3. The process of Claim 1 wherein step (c) fractionation of said C₃ + hydrocarbon stream comprises passing said stream to a depropanizer; separating an overhead stream rich in C₃ hydrocarbons, a bottom stream rich in C₅ + gasoline range liquid and a stream containing C₄ hydrocarbons; passing said C₄ hydrocarbon stream to a stripper for separation of a bottom stream therefrom rich in C₄ hydrocarbons and recycling said stripper overhead stream to said depropanizer.
4. The process of Claim 1 wherein said lower olefinic hydrocarbon feedstock comprises the overhead effluent from fluid catalytic cracking (FCC) main column overhead product debutanizer or depropanizer.
5. The process of Claim 1 wherein step (c) fractionation comprises depropanizing to produce a C₄ + hydrocarbon stream; passing said stream to an FCC debutanizer for separation; recovering a bottom stream comprising C₅ + gasoline range hydrocarbon and an overhead stream comprising C₄ hydrocarbons.
6. The process of Claim 1 wherein said solid catalyst comprises zeolite type metallosilicate.
7. The process of Claim 6 wherein said zeolite type metallosilicate comprises ZSM-5.
8. The process of Claim 2 which comprises separating oligomerization reaction effluent in a low temperature separator; passing the gaseous stream from the separator to the absorber means for contact with liquid hydrocarbons to recover C₂-hydrocarbons; recovering a liquid stream from the low temperature separator comprising volatile C₅ + gasoline; stripping the low temperature separator liquid stream to recover a bottom stripping fraction comprising C₃ + hydrocarbons; passing said bottom stripping fraction to a fractionator system for separation of an overhead stream comprising C₃ hydrocarbons, a bottom stream comprising C₅ + gasoline liquids and a stream comprising C₄

hydrocarbons.

9. The process of claim 4 wherein bottoms fraction of said FCC depropanizer is passed to debutanizer in combination with C₄ + fraction from said oligomerization zone.

Patentansprüche

1. Verbessertes Verfahren zur Umwandlung eines niederen olefinischen Kohlenwasserstoffbeschickungs-
materials in Kohlenwasserstoffe im C₅ +-Benzinbereich, welches umfaßt:
 - a. Kontakt eines Kohlenwasserstoffstroms, der olefinische C₃ -- und/oder C₄ --Kohlenwasserstoffe
enthält, mit einem formselektiven festen Katalysator mit mittleren Poren in einer Oligomerisierungs-
zone bei Oligomerisierungsbedingungen, wodurch ein Abflußstrom hergestellt wird, der reich an
Kohlenwasserstoffen in C₅ +-Benzinbereich ist;
 - b. Abtrennung dieses Abflußstroms, wodurch ein C₃ --Kohlenwasserstoffstrom und ein C₃ +-Kohlen-
wasserstoffstrom bereitgestellt werden;
 - c. Fraktionieren des C₃ +-Kohlenwasserstoffstroms, wodurch ein Kohlenwasserstoffstrom in C₅ +-
Benzinbereich, ein an C₄-Kohlenwasserstoffen reicher Strom und ein an C₃-Kohlenwasserstoffen
reicher Strom bereitgestellt werden.
2. Verfahren nach Anspruch 1, wobei der Schritt (b) der Abtrennung des Abflußstromes umfaßt: Leiten
dieses Stromes zu Hochtemperatur- und/oder Niedertemperatur-Separatoren; Leiten des gasförmigen
Anteils aus diesen Separatoren zu einer Absorbereinrichtung, wodurch ein C₂ --Kohlenwasserstoffstrom
erzeugt wird; und Leiten des flüssigen Anteils aus den Separatoren zu einer Stripping-Einrichtung,
wodurch ein C₃ +-Kohlenwasserstoffstrom erzeugt wird.
3. Verfahren nach Anspruch 1, wobei der Schritt (c) der Fraktionierung des C₃ +-Kohlenwasserstoffstroms
umfaßt: das Leiten des Stroms zu einem Depropaner; die Abtrennung eines Kopfproduktstroms, der
reich an C₃-Kohlenwasserstoffen ist, eines Rückstandsstroms, der reich an einer Flüssigkeit im C₅ +-
Benzinbereich ist, und eines Stroms der C₄-Kohlenwasserstoffe enthält; das Leiten des C₄-Kohlenwas-
serstoffstroms zu einem Stripper, damit davon ein Rückstandsstrom abgetrennt wird, der reich an C₄-
Kohlenwasserstoffen ist, und die Rückführung dieses Kopfproduktstroms vom Stripper zum Depropa-
ner.
4. Verfahren nach Anspruch 1, wobei das niedere olefinische Kohlenwasserstoffbeschickungsmaterial den
Kopfproduktabfluß eines Debutaners oder Depropaners für das Kopfprodukt der Hauptkolonne für das
katalytische Wirbelschichtcracken (FCC) umfaßt.
5. Verfahren nach Anspruch 1, wobei der Schritt (c) der Fraktionierung umfaßt: Depropanisieren, wodurch
ein C₄ +-Kohlenwasserstoffstrom erzeugt wird; Leiten des Stroms zu einem FCC-Debutaner zur Antren-
nung; Gewinnung eines Rückstandsstroms, der C₅ +-Kohlenwasserstoffe im Benzinbereich umfaßt, und
eines Kopfproduktstroms, der C₄-Kohlenwasserstoffe umfaßt.
6. Verfahren nach Anspruch 1, wobei der feste Katalysator einen Metallosilicat vom Zeolith-Typ umfaßt.
7. Verfahren nach Anspruch 1, wobei der Metallosilicat vom Zeolith-Typ ZSM-5 umfaßt.
8. Verfahren nach Anspruch 2, welches umfaßt: Antrennung des Abflusses der Oligomerisierungsreaktion
in einem Niedertemperatur-Separator; Leiten des gasförmigen Stroms aus dem Separator zu einer
Absorbereinrichtung zum Kontakt mit flüssigen Kohlenwasserstoffen, wodurch C₂ --Kohlenwasserstoffe
gewonnen werden; Gewinnung eines flüssigen Stroms aus dem Niedertemperatur-Separator, der
flüchtiges C₅ +-Benzin umfaßt; Strippen des flüssigen Stroms des Niedertemperatur-Separators, wo-
durch eine Rückstandsfraktion vom Strippen gewonnen wird, die C₃ +-Kohlenwasserstoffe umfaßt; Leiten
der Rückstandsfraktion vom Strippen zu einem Fraktionatorsystem zur Antrennung eines Kopfprodukt-
stromes, der C₃-Kohlenwasserstoffe umfaßt, eines Rückstandsstroms, der C₅ +-Benzinflüssigkeiten
umfaßt, und eines Stroms, der C₄-Kohlenwasserstoffe umfaßt.
9. Verfahren nach Anspruch 4, wobei die Rückstandsfraktion des FCC-Depropaners zusammen mit der
C₄ +-Fraktion aus der Oligomerisierungszone zum Debutaner geleitet wird.

Revendications

1. Un procédé amélioré de conversion de charge d'hydrocarbures oléfiniques en hydrocarbures de la gamme d'ébullition des essences en $C_5 +$, comprenant les étapes suivantes:
 - 5 a) mise d'un courant d'hydrocarbures contenant des hydrocarbures oléfiniques en C_3- et/ou C_4 au contact d'un catalyseur solide sélectif de forme de pores moyens, dans une zone d'oligomérisation dans des conditions d'oligomérisation permettant d'obtenir un courant effluent riche en hydrocarbures de la gamme d'ébullition des essences en $C_5 +$;
 - 10 b) séparation dudit courant effluent pour obtenir un courant d'hydrocarbures en C_3 et un courant d'hydrocarbures en $C_3 +$;
 - c) fractionnement du courant d'hydrocarbures en $C_3 +$ pour obtenir un courant d'hydrocarbures de la gamme d'ébullition des essences en $C_5 +$, un courant riche en hydrocarbures en C_4 et un courant riche en hydrocarbures en C_3 .
- 15 2. Le procédé selon la revendication 1, dans lequel l'étape (b) de séparation dudit courant effluent consiste à faire passer ledit courant dans des séparateurs à haute température et/ou à basse température, à faire passer la portion gazeuse desdits séparateurs dans des absorbeurs dans lesquels est obtenu le courant d'hydrocarbures en C_2- ; et à faire passer la portion liquide provenant de ces séparateurs dans des extracteurs desquels est retiré le courant d'hydrocarbures en $C_3 +$.
- 20 3. Le procédé selon la revendication 1, dans lequel l'étape (c) de fractionnement de ce courant d'hydrocarbures en $C_3 +$ consiste à faire passer ledit courant dans un dépropaniseur; à séparer un courant de tête riche en hydrocarbures en $C_3 +$, un courant de cuve riche en liquide de la gamme d'ébullition des essences et un courant contenant des hydrocarbures en C_4 ; à faire passer ledit courant d'hydrocarbures en C_4 dans un extracteur pour la séparation d'un courant de cuve riche en hydrocarbures en C_4 et à recycler ledit courant de tête de l'extracteur vers ledit dépropaniseur.
- 25 4. Le procédé selon la revendication 1, dans lequel ladite charge d'hydrocarbures oléfiniques inférieures comprend l'effluent de tête provenant du débutaniseur ou dépropaniseur de produit de tête de colonne principale FCC (craquage catalytique en lit fluidisé).
- 30 5. Le procédé selon la revendication 1, dans lequel l'étape (c) de fractionnement consiste à dépropaniser pour obtenir un courant d'hydrocarbures en $C_4 +$, à faire passer ledit courant dans un débutaniseur FCC pour une séparation, à récupérer un courant de cuve comprenant des hydrocarbures de la gamme d'ébullition des essences en $C_5 +$ et un courant de tête comprenant des hydrocarbures en C_4 .
- 35 6. Le procédé selon la revendication 1, dans lequel ledit catalyseur solide comprend un métalosilicate de type zéolitique.
- 40 7. Le procédé selon la revendication 1, dans lequel ledit métalosilicate de type zéolitique comprend la ZSM-5.
- 45 8. Le procédé selon la revendication 2, qui consiste à séparer un effluent de réaction d'oligomérisation dans un séparateur à basse température; à faire passer le courant gazeux provenant du séparateur dans des absorbeurs pour contact avec des hydrocarbures liquides pour récupérer les hydrocarbures en C_2 ; à récupérer un courant liquide sortant du séparateur à basse température comprenant des essences volatiles en $C_5 +$; à extraire le courant liquide du séparateur à basse température pour récupérer une fraction d'extraction de cuve comprenant des hydrocarbures en $C_3 +$; à faire passer cette fraction d'extraction de cuve dans un système de fractionnement pour obtenir un courant de tête comprenant des hydrocarbures en C_3 , un courant de cuve comprenant des liquides de la gamme d'ébullition des essences en $C_5 +$ et un courant contenant des hydrocarbures en C_4 .
- 50 9. Le procédé selon la revendication 4, dans lequel on fait passer la fraction de cuve dudit dépropaniseur FCC à un débutaniseur en association avec la fraction en $C_4 +$ provenant de ladite zone d'oligomérisation.
- 55

FIG. 1

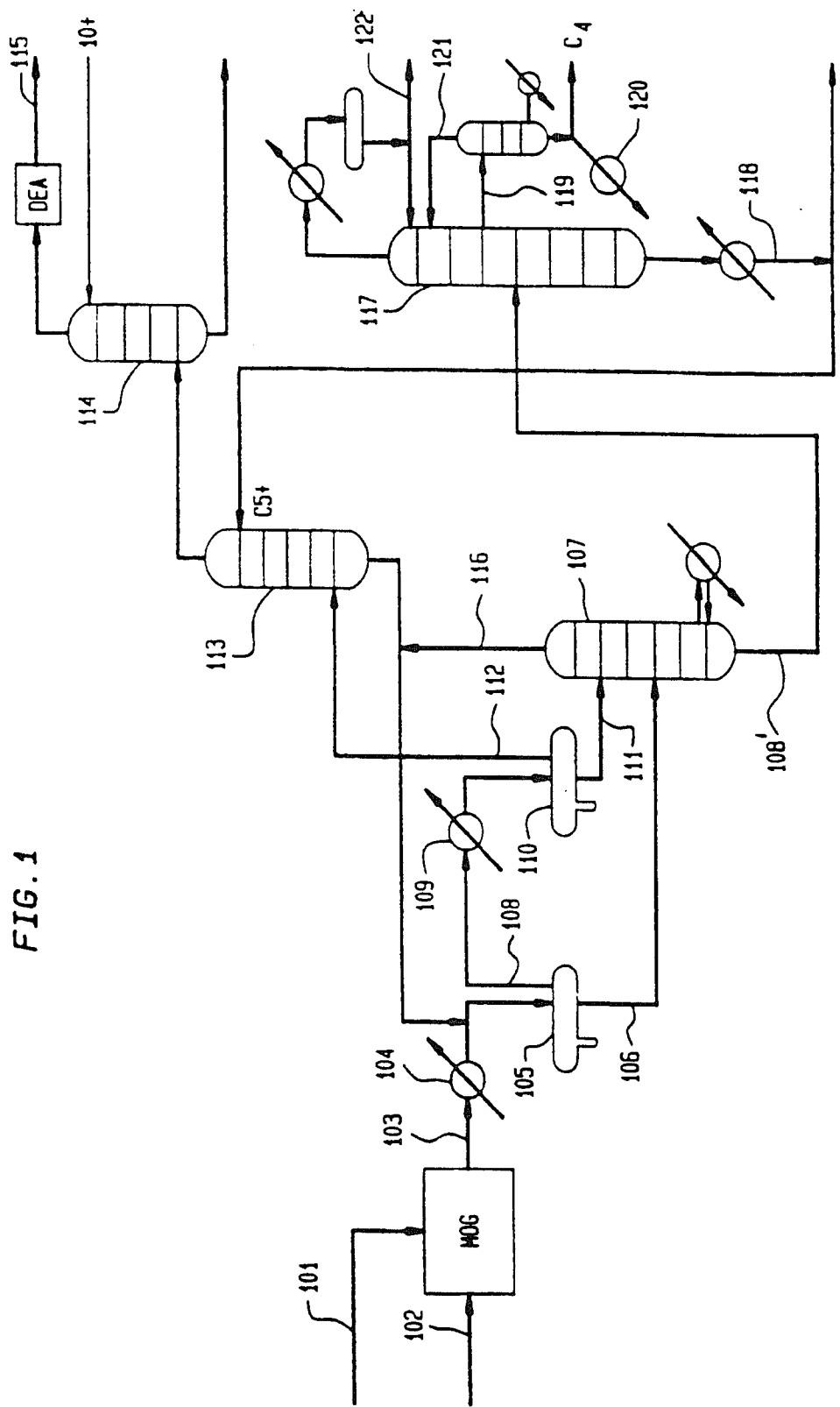


FIG. 2

